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DESIGN CRITERIA FOR A PACKED-FLUIDIZED BED: HOMOGENEOUS FLUIDIZATION OF GELDART'S CLASS B SOLIDS

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ABSTRACT
Homogeneous fluidization of Geldart's group B solids was investigated in two packed columns (5 and 10 cm OD) whose packing spheres were either 0.41 cm lead shots or 1.1 cm glass beads. Some criteria to obtain regular regimes of homogeneous fluidization were fixed and the parameters of the Richardson-Zaki equation which describes the expansion process were identified.

INTRODUCTION
The interstitial network provided by a packed solid constitutes a confining environment in which a finer solid can be suspended without the formation of bubbles. Thus, homogeneous fluidization of Geldart's group B particles can be obtained if they are let expand in the voids of a packed bed (1-7). This method seems likely to provide a very efficient fluid-solid contact mode, suitable for high conversion of gaseous reactants or nearly complete adsorption of specific components of the fluidizing stream. Operations of this kind could perhaps be conducted in a packed-fluidized bed over a broad velocity range without the handling problems orderly associated to the use of fine powders. That is even more true for applications in which the solid of interest (for instance a sorbent or a catalyst) is a powder obtained from a synthesis, then granulated in nearly spherical pellets of larger size. What reported in the present work shows how to assign to the main variables that influence the behaviour of a packed-fluidized bed values suitable for obtaining a regular regime of homogeneous fluidization.

EXPERIMENTAL
In order to explore the potentiality of the technique, various experiments were carried out in two columns with an outer diameter of 5 cm (ID=4.95 cm) and 10 cm (ID=9.32 cm), respectively. In all of them, both the confining packed bed and the particles subjected to fluidization with air in its void matrix were approximately spherical. Two different types of packing were used, as either 0.41 cm lead spheres or 1.1 cm glass beads were used. The fluidized solids were several cuts of glass ballotini, whose average size d_f ranged from 100 to 588 μm, as well as samples of ceramic (CE), zirconium oxide (ZO) and bronze spheres (BR) (see Tables 1 and 2). The height H_{fc} of the finer bed in the fixed state was orderly 5 cm, although additional experiments with heights up to 25 cm were also carried out to verify the invariance of the minimum fluidization velocity with the solid mass. The height of the packed bed was accordingly set equal to 4 H_{fc}, to explore the whole homogeneous expansion regime of each solid sample.

In a typical experiment the packing spheres are first poured onto the column to form a packed bed of height H_p; subsequently, the finer solid is loaded and its height H_{fc} recorded after a complete fluidization-defluidization cycle. At the end of these preliminary steps, the bed configuration is that illustrated in Fig.1.

$$\varepsilon_p = 1 - \frac{m_p}{\rho_p\, A\, H_p} = 1 - \alpha_p \quad (1)$$

$$\alpha_f = \frac{m_f}{\rho_f\, A\, H_{fc}} \quad (2)$$

$$\varepsilon = 1 - \alpha_p - \alpha_f \quad (3)$$

$$\varepsilon_{fc} = \frac{\varepsilon}{\varepsilon_p} \quad (4)$$

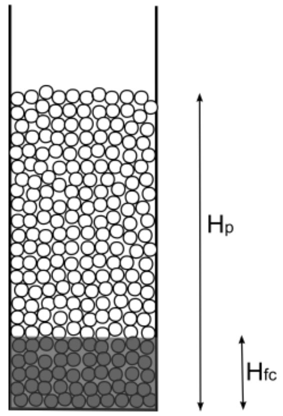

Fig.1 - Sketch of the confined fluidized bed.

After measuring the heights of the two beds have been measured, the packing bed voidage ε_p, the volumetric solid fractions (α_p, α_f), the bed voidage ε and the equivalent voidage ε_{fc} (i.e. the voidage that the fine bed would have in a column of section $A\varepsilon_p$) are calculated according to equations (1)-(4). The experiment is then started and the gas flow velocity progressively increased; the total pressure drop and the bed height are determined at each value of u until the expansion of the fine solid brings it to reach the free surface of the coarse packing.

Table 1. Experimental results and calculated parameters. Column diameter: 5 cm; Packing diameter: 0.41 cm

	Particle density ρ_f [g cm^{-3}]	Sauter diameter d_f [10^{-4}cm]	d_f/d_h [-]	Confined u_{mfc} [cm s^{-1}]	n_{exp} [-]	$u_{0,exp}$ [cm/s]	n_{calc} [-]	$u_{0,calc}$ [cm/s]
GB 90-125 μm	2.48	100	0.060	0.50	4.27	89	4.26	90
GB 125-150 μm		136	0.082	0.83	4.08	107	4.05	116
GB 150-200 μm		171	0.103	1.10	4.03	120	3.93	134
GB 200-250 μm		228	0.137	2.48	3.98	149	3.85	152
GB 250-300 μm		271	0.163	2.95	3.78	161	3.80	161
GB 300-355 μm		327	0.197	4.88	3.60	173	3.79	169
GB 355-400 μm		361	0.218	6.26	3.61	189	3.79	172
CE 200-250 μm	3.78	230	0.139	1.55	3.56	186	3.72	196
CE 250-300 μm		268	0.161	2.58	3.90	234	3.70	204
ZO 200-250 μm	6.15	245	0.148	4.38	3.70	260	3.58	262
ZO 250-300 μm		261	0.157	6.32	3.52	263	3.60	267
BR 200-250 μm	8.75	229	0.138	4.66	3.56	321	3.53	314
BR 250-300 μm		272	0.164	8.99	3.26	339	3.48	327

GB=glass ballotini, CE=ceramic, ZO=zirconium oxide, BR=bronze, LS=lead spheres

Table 2. Experimental results and calculated parameters.
Column diameter: 10 cm; Packing diameter: 1.1 cm

	Particle density ρ_f [g cm^{-3}]	Sauter diameter d_f [10^{-4}cm]	d_f/d_h [-]	Confined u_{mfc} [cm s^{-1}]	n_{exp} [-]	$u_{0,exp}$ [cm/s]	n_{calc} [-]	$u_{0,calc}$ [cm/s]
GB 300-355 μm	2.48	319	0.070	3.85	3.30	241	3.31	252
GB 400-500 μm		460	0.102	8.54	3.11	303	3.19	282
GB 560-630 μm		588	0.130	12.5	2.83	324	3.15	294

RESULTS

The two diagrams of Fig.2 illustrate the typical variation of the pressure drop across the confined bed, ΔP_{conf}, and that of the bed voidage ε versus the superficial gas velocity. ΔP_{conf} is the difference between the total pressure drop and that relevant to the upper region of the packing devoid of fines:

$$\Delta P_{conf} = \Delta P_{tot} - \Delta P_{free\ packing} \tag{5}$$

Data of Fig.2 were obtained in the smaller column, with GB136 and BR229 as confined particles, and reveal a different behaviour in the particulate fluidization of the two systems. Bronze particles exhibit more regular homogeneous expansion, which starts at a definite value of u_{mfc}. Voidage variation is described by a unique power law and the corresponding pressure drop is described by Ergun's equation. On the other hand, glass particles do not show an equally sharp transition to homogeneous fluidization; moreover, their expansion follows two different trends: at relatively low velocities the pressure drop tends to remain constant as in conventional fluidization, whereas a monotonic increase is observed at higher velocities.

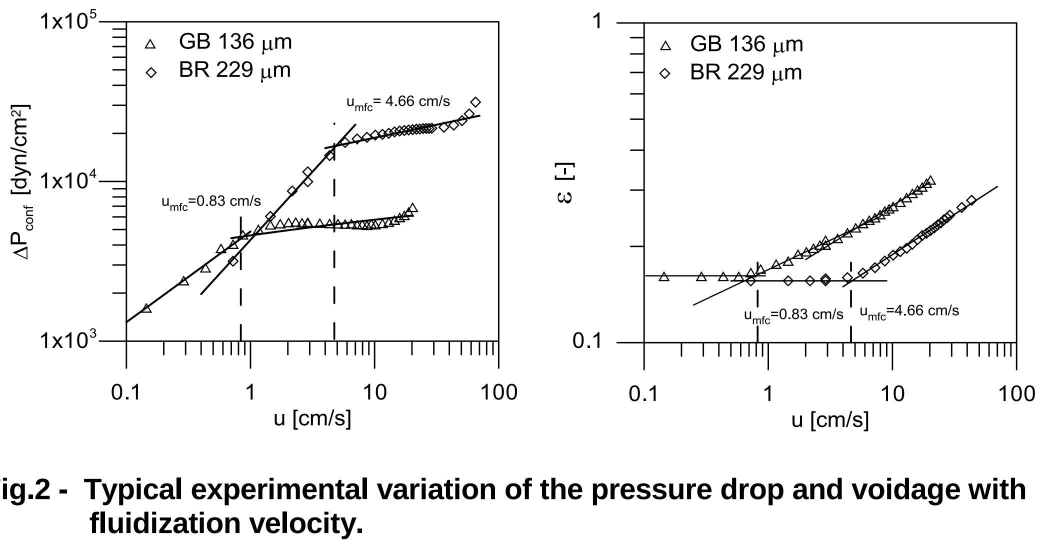

Fig.2 - Typical experimental variation of the pressure drop and voidage with fluidization velocity.

Inspection of this type of diagrams for all the solids investigated suggests a criterion for selecting the diameter of the packed particles as suitable for the specific solid to be fluidized in their voids. As discussed elsewhere (8), a geometrical parameter capable to describe the interaction between solid is the equivalent hydraulic diameter of the void network of the packed bed

$$d_h = \frac{2}{3}\frac{\varepsilon_p}{(1-\varepsilon_p)}\, d_p \qquad (6)$$

whose value for the two systems of Fig.2 results equal to 0.168 and 0.453 cm, respectively. In more general terms, it can be stated that solids with a ratio d_f/d_h lower than about 0.1 display an anomalous homogeneous regime. This is the case of the three finest cuts of GB in the 5 cm column and of the two finest cuts in the 10 cm column. A possible interpretation of this finding is that in the fixed state small particles can penetrate the voids of the packing more deeply than bigger ones, up to the region close to the contact points between coarse spheres. It may be thought, therefore, that at relatively low velocities the amount of particles involved in the expansion process is somewhat limited and that higher velocities are required to suspend the whole fine mass.
In the light of these considerations, the following criteria are suggested to obtain a homogeneous regime of confined fluidization: d_h should not be larger than $10d_f$; to ensure smooth percolation, the choice of the packing diameter should be such that $d_f/d_p \leq 0.15$. In addition, the ratio D/d_p has to be not lower than 5, to ensure negligible effects of the column wall (9). Since with packings of spheres ε_p is approximately equal to 0.38, eq. (7) sets the maximum value of d_p as

$$d_{p,max} = \frac{3}{2}\frac{(1-\varepsilon_p)}{\varepsilon_p}\, d_h \cong 2.5\, d_h \cong 25\, d_f \qquad (7)$$

so that, given d_f, the advisable range of variation of d_p is determined as

$$d_{p,min} = \frac{d_f}{0.15} \leq\, d_p \leq d_{p,max} = 25\, d_f \qquad (8)$$

An additional requirement originates from the necessity of minimizing the relative weight of the fluid dynamic singularities typical of the distributor region on the overall behaviour of the confined bed. Experimental measurements of u_{mfc} carried out on beds of different height, give rise to the typical trends of Fig.3; it is then verified that u_{mfc} becomes independent of the confined bed mass at $H_{fc}/D \geq 2$. These constant values of u_{mfc} are reported in Tabs 1 and 2.

As regards the packed bed height H_p, it has to be set at values that allow the desired level of voidage, i.e the desired degree of expansion of the confined bed. This result can be accomplished by means of the set of equations (1)-(4). Consistently with what reported in other studies (7, 8) the process of expansion of confined gas-fluidized beds is regulated by a modified form of the Richardson and Zaki's equation. As first proposed by Glasserman et al. (6), the parameters n and u_0 can be obtained by fitting the experimental values of u/ε_p versus ε_{fc} with a straight line on a logarithmic plot.

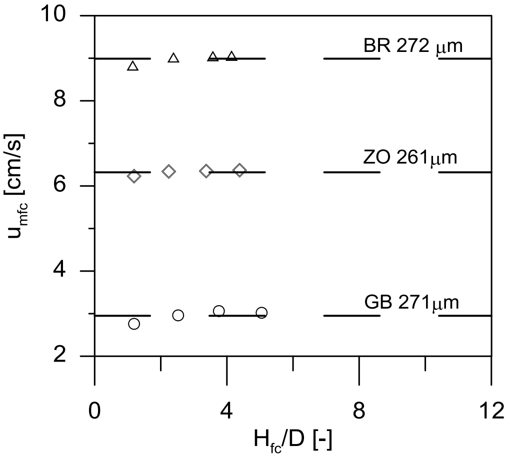

Fig.3 - Variation of the confined fluidization velocity with the bed aspect ratio.

The expansion trends relevant to some of the materials of the present study are shown in the three diagrams of Fig.4. The values of n and u_0 obtained by this procedure are reported in Tabs 1 and 2.

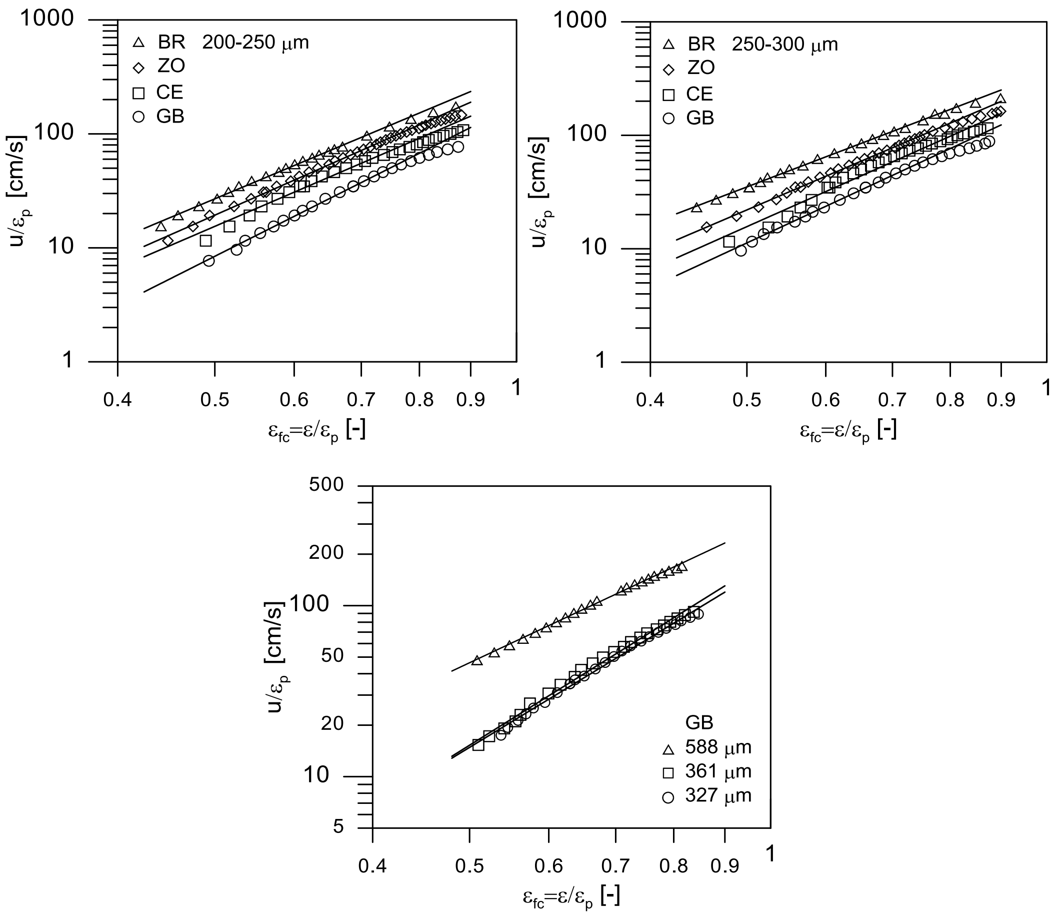

Fig.4 - Expansion diagrams of the confined fluidized beds.

THEORY

Whenever regular homogeneous expansion is observed, the exponent n of the Richardson and Zaki's equation can be calculated by the relationship reported in a previous work (8):

$$n = \left(4.45 + 5.74\frac{d_f}{d_h}\right)\frac{1}{Re_t^{0.1}} \tag{9}$$

The Reynolds number at terminal velocity, Re_t, is related to the Archimedes number by the following equation, reported by Gibilaro (10):

$$Re_t = [-3.809 + (3.809^2 + 1.832\,Ar^{0.5})^{0.5}]^2 \tag{10}$$

where $$Ar = \frac{g\,d_f^3(\rho_f - \rho_g)\rho_g}{\mu_g^2} \quad \text{and} \quad Re_t = \frac{\rho_g\,d_f\,u_t}{\mu_g} \tag{11}$$

The values of n calculated from eq. (9) are reported in Tabs 1 and 2. As shown in Fig.5, where a comparison is made with the experimental values of the same parameter, the agreement is satisfactory in almost all cases. A simple model for estimating u_0, i.e. the maximum expansion velocity of the confined bed, is provided by a force balance on the single particle, whose general one-dimensional form is that reported by Wallis (11). Under the assumption of uniform steady flow and negligible contact forces, it becomes

$$-V\frac{dp}{dz} + F_d - V\rho_f\,g = 0 \tag{12}$$

where V is the single particle volume and F_d is the drag force that acts on it. Conceptually, in a fluidized bed the total pressure gradient (which includes the static pressure drop) is $[(1-\varepsilon)\rho_f+\varepsilon\rho_g]g$; thus, eq. (12) yields

$$F_d = V(\rho_f - \rho_g)g\varepsilon \tag{13}$$

Since at ε=1, F_d must be equal to the drag force F_{dt} at terminal conditions, the following relationship, coincident with that reported by Di Felice (12), is obtained:

$$F_d = F_{dt}\,\varepsilon \tag{14}$$

As far as the confined system is regarded as a fluidized bed of cross-sectional area $A\varepsilon_p$, eq. (14) should still be valid. F_{dt} can then be calculated from a suitable correlation for the friction factor. To this purpose, a good choice seems that of the equation derived from Dallavalle by Gibilaro (10) as its simple form covers the whole field of variation of the Reynolds number. Given that at the maximum expansion velocity $\varepsilon=\varepsilon_p$ (so that ε_{fc}=1), this substitution changes Eq. (14) into:

$$F_d = \left(0.63 + \frac{4.8}{Re_0^{0.5}}\right)^2\left(\frac{1}{2}\rho_g\,u_0^2\right)\left(\frac{\pi\,d_f^2}{4}\right)\varepsilon_p \tag{15}$$

where $Re_0=\rho_g u_0 d_f/\mu_g$. Eq. (15) gives the drag force on the single particle in the absence of packing, i.e. when ε_p=1. When the fine particles are extremely diluted inside the packed bed, the pressure gradient is that due only to the coarse spheres. In a fixed bed of cross-sectional area $A\varepsilon_p$ it can thus be calculated as

$$-\frac{dp}{dz} = 150\,\frac{\mu_g\,(1-\varepsilon_p)^2}{\varepsilon_p^3\,d_p^2}\,u_0 + 1.75\,\frac{\rho_g\,(1-\varepsilon_p)}{\varepsilon_p^3\,d_p}\,u_0^2 \qquad (16)$$

Substitution of equations (15) and (16) into (12) allows calculating the maximum expansion velocity u_0. As done for the parameter n, the calculated values of u_0 are reported in Tabs 1 and 2. Again, Fig.5 shows that the error associated with this method of prediction is acceptable, as it seldom exceeds 10%.

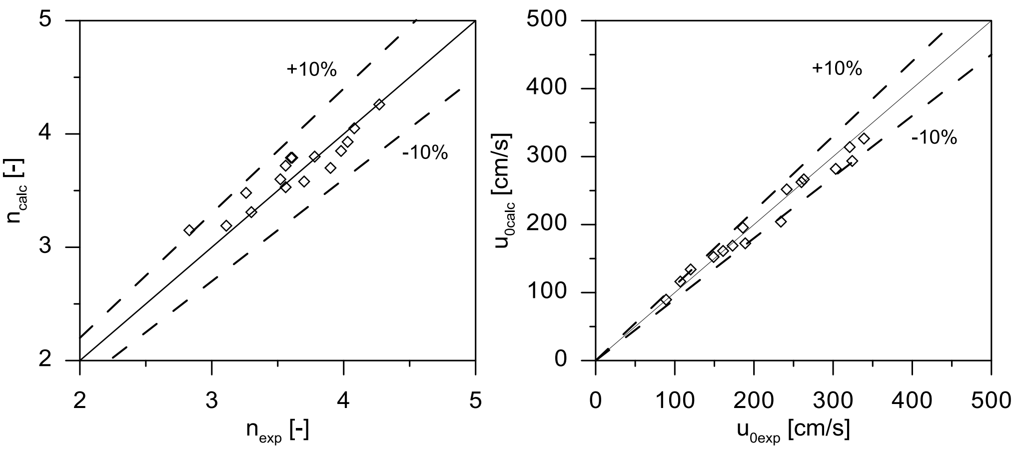

Fig.5 - Comparison between experimental and calculated values of the expansion index and of the maximum expansion velocity of confined beds.

CONCLUSIONS

The results reported in this work allow fixing some criteria to obtain regular regimes of homogeneous fluidization. They regard: the choice of the packed bed particle size; the minimum aspect ratio of the confined bed that guarantees the minimum fluidization velocity to be independent of the solid mass; the packed bed height necessary to operate the particle system over a broad field of homogeneous expansion.

As far as these criteria are obeyed, the homogeneous expansion of Geldart's B particles can be described by a modified form of the Richardson-Zaki equation. The two parameters of this relationship, namely the expansion index n and the maximum expansion velocity u_0, are predicted with good accuracy by the model proposed.

NOTATION

A	column cross-sectional area, cm^2	u_{mfc}	minimum fluidization velocity of the confined bed, cm/s
Ar	Archimedes Number ($=gd_p^3(\rho_f-\rho_g)\rho_g/\mu_g^2$), -	u_t	terminal velocity, cm/s

D	column internal diameter, cm	u_0	maximum expansion velocity in the packed bed, cm/s
d_f	fine solid diameter, cm	V	fine particle volume, cm^3
d_h	hydraulic diameter of the voids , cm	z	vertical distance above the distributor, cm
d_p	diameter of the packed solid, cm		
F_d	Drag force on the single particle, dyn	**Greek symbols**	
F_{dt}	Drag force on the single particle at terminal conditions, dyn	$\alpha_{f/p}$	fraction of the fine /packed solid, -
H_{fc}	confined bed height, cm	ΔP_{conf}	pressure drop in the confined system, dyn/cm^2
H_p	packed bed height, cm	ε	voidage of the confined bed, -
n	expansion index, -	ε_{fc}	voidage of the equivalent conventional bed ($=\varepsilon/\varepsilon_p$), -
P	pressure, dyn/cm^2	ε_p	voidage of the packed bed, -
Re_t	terminal Reynolds Number ($=\rho_g u_t d_f/\mu_g$), -	μ	gas viscosity, (g/cm s)
Re_0	Reynolds Number at the maximum expansion velocity ($=\rho_g u_0 d_f/\mu_g$), -	ρ_f	density of the fine solid, g/cm^3
u	superficial gas velocity, cm/s	ρ_g	gas density, g/cm^3

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